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Economic and Technical Evaluation of the Sulfurous Acid-Caustic Purification Process for Producing Alumina From Kaolinitic Clay

By Deborah A. Kramer



UNITED STATES DEPARTMENT OF THE INTERIOR

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UNIT OF MEASURE ABBREVIATIONS USED IN THIS REPORT

Btu	British thermal unit	gal/ft ² ·h	gallon per square foot per hour
Btu/d	British thermal unit per day	h	hour
Btu/gal	British thermal unit per gallon	h/d	hour per day
Btu/lb	British thermal unit per pound	kW·h	kilowatt-hour
°C	degree Celsius	lb	pound
d/wk	day per week	Mgal	thousand gallon
d/yr	day per year	min	minute
°F	degree Fahrenheit	MMBtu	million British thermal unit
ft	foot	pct	percent
ft ²	square foot	psig	pound per square inch, gauge
gal	gallon	ton/d	short ton per day
gal/d	gallon per day	yr	year

ECONOMIC AND TECHNICAL EVALUATION OF THE SULFUROUS ACID-CAUSTIC PURIFICATION PROCESS FOR PRODUCING ALUMINA FROM KAOLINITIC CLAY

By Deborah A. Kramer¹

ABSTRACT

An economic and technical evaluation of a proposed process to recover alumina by leaching calcined kaolinitic clay with sulfurous acid is presented in this Bureau of Mines report. After the insoluble portion of the clay has been removed by leaching, monobasic aluminum sulfite is precipitated from the pregnant leach solution by partial hydrothermal decomposition. Further heating completes the decomposition to an impure alumina hydrate. This intermediate is purified in a modified Bayer process, which includes dissolving the alumina hydrate in a caustic solution at high temperature and pressure, then decreasing the temperature and pressure to precipitate alumina trihydrate. Calcining the trihydrate produces the alumina product.

A cost estimate was prepared for a plant producing 1,000 tons of alumina per day, 350 d/yr. The estimated operating cost is approximately \$475 per ton of alumina. This is substantially higher than a comparable estimated operating cost for the Bayer process of \$250 per ton of alumina. The proposed process does not appear economically attractive, even as a method to modify a current Bayer processing plant to use a kaolinitic clay feed. High capital costs and high energy requirements make the process prohibitively expensive.

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INTRODUCTION

The Bureau of Mines is investigating technologies for the recovery of alumina from domestic nonbauxitic resources. Most of the alumina consumed in the United States is produced from imported bauxite. The United States has very limited supplies of bauxite, but has large reserves of alumina-containing clay and other resources. Development of an economic process to recover the alumina from kaolinitic clay or other domestic resources would insure a continuing alumina supply to meet national needs.

The Bureau of Mines has previously evaluated a number of processes for recovering alumina from clay and anorthosite. None of these processes appeared economically competitive with the Bayer process; however, a few warranted further investigation because their costs were not appreciably higher than those of the Bayer process. The technology developed by these investigations could prove significant if the cost of imported bauxite increases.

One of these processes involves sulfurous acid leaching of kaolinitic clay, then purification of the crude alumina product by a modified Bayer process. Two previous evaluations of the sulfurous acid process, based on limited data,

indicated that the proposed process appeared economically attractive (1-2).²

The sulfurous acid process dates back to the late 1800's. Development was attributed to Th. Goldschmidt A.G. This process occasioned numerous patents (3-10). In 1938, VAWAG at Lippewerk, Germany, tried the Goldschmidt process on a commercial scale but found that the alumina product contained an excessive amount of impurities. It was then that Bayer purification was added to produce a cell-grade alumina (11).

In the early 1980's researchers at the Bureau of Mines investigated the sulfurous acid leaching process on a laboratory scale. The objective of this research was to expand the experimental data base on the process, to confirm assumptions used in previous evaluations, and to determine if additional investigations in the Bureau's miniplant project were warranted (12).

The present study includes data (poor alumina recoveries, longer leaching time, etc.) from the recent Bureau research (12) and has been prepared to reassess the potential economics and compare the results with the previous evaluations.

PROCESS DESCRIPTION

In the proposed process, raw kaolinitic clay is calcined to convert the contained alumina to an acid-soluble form. The calcined clay is leached with a sulfurous acid solution. Alumina is extracted in the form of aluminum sulfite, which is separated from the insoluble residue by filtration. The temperature and pressure of the pregnant leach solution are increased to precipitate monobasic aluminum sulfite. Decomposition of the monobasic aluminum sulfite produces an impure alumina hydrate.

The impure alumina hydrate is purified in a modified Bayer process. First, it is dissolved in a caustic solution at

high temperature and pressure. Undissolved impurities are removed, and alumina trihydrate is precipitated by cooling the solution to the point of supersaturation, then seeding with fine trihydrate crystals. The trihydrate is calcined to obtain the final alumina product. A basic process flowsheet is shown in figures 1 and 2 for a plant designed to produce 1,000 tons of alumina per day based on 3 shifts per day operation, 350 d/yr. For purposes of discussion, the plant is divided into seven major

²Underlined numbers in parentheses refer to items in the list of references preceding the appendix.

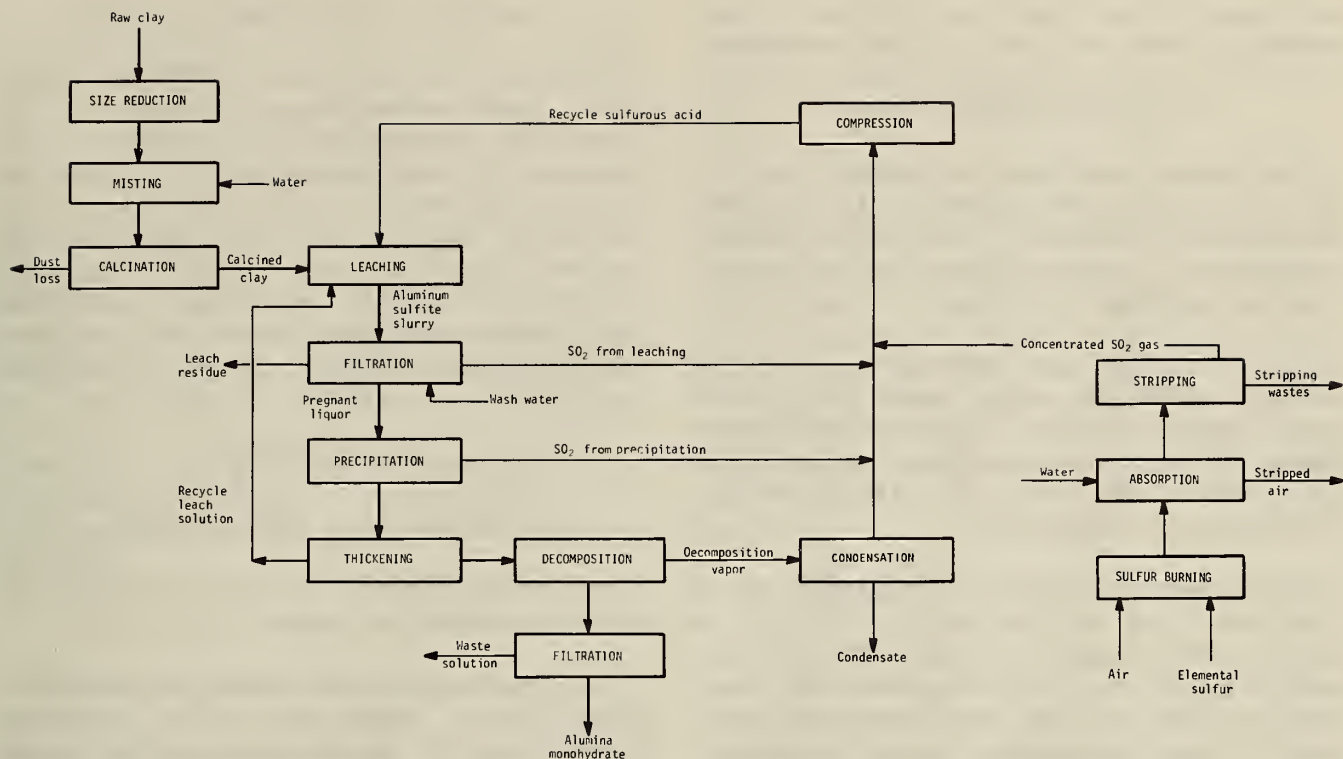


FIGURE 1. - Sulfurous acid digestion and crude alumina production.

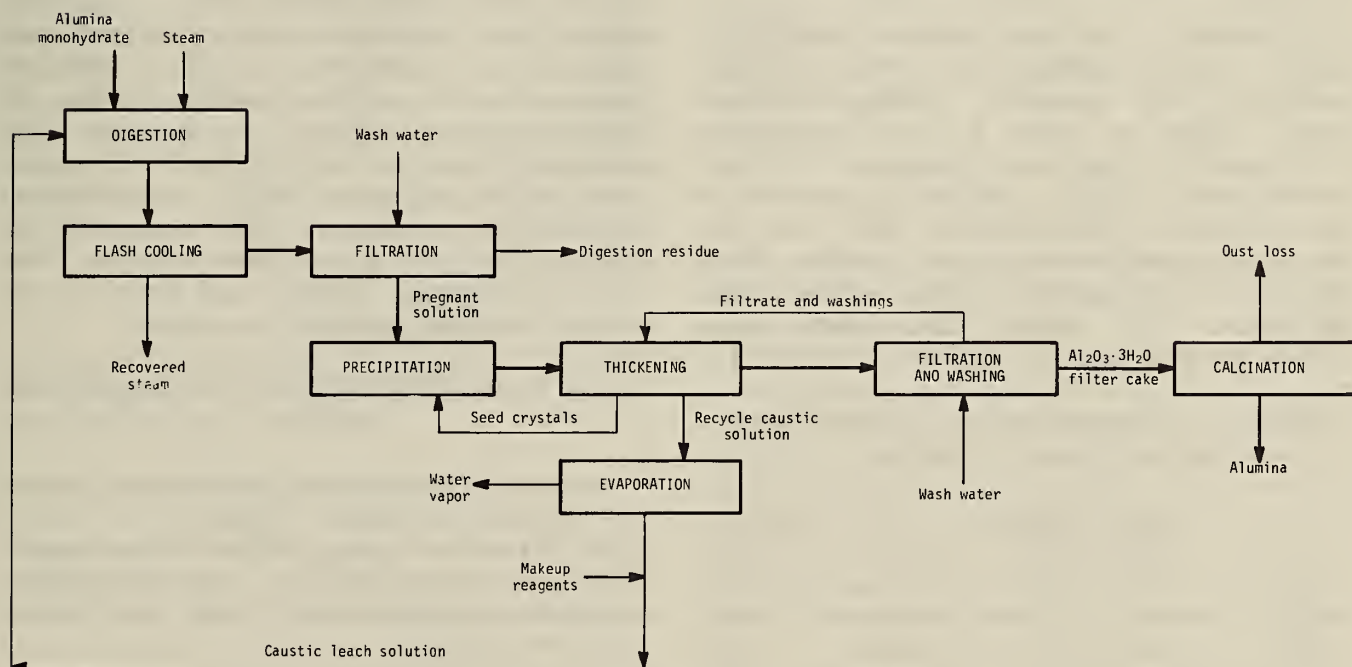


FIGURE 2. - Caustic purification.

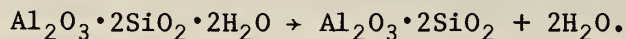
sections. The material balances for these sections are shown in the appendix.

CLAY PREPARATION SECTION

Kaolinitic clay is delivered to the plant by truck from a nearby clay pit and dumped into a hopper. The clay is then conveyed to a covered storage pile containing a 30-day supply.

Clay is withdrawn from this pile as needed and fed to hammer mills for crushing to minus 16 mesh. Discharge from the hammer mills is conveyed to a pelletizing disk, where the clay is misted before being fed to surge bins.

Minus 16-mesh clay is calcined in a fluidized bed at 750° C (1,382° F) to remove the 18 pct free moisture and the combined water. The calcination reaction, which converts the contained alumina to an acid-soluble form, is assumed to be



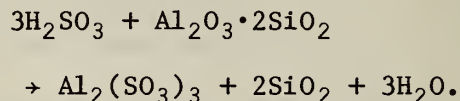
Calcined clay is cooled to 65° C (150° F) and conveyed to the leaching section. The analysis of the calcined clay, shown in table 1, was taken from the material balance prepared by the researchers (12). The SiO₂ content of the calcined clay reported by the researchers was assumed to include TiO₂ known to be present in the clay. The estimated TiO₂ content is shown separately in the analysis given in table 1.

TABLE 1. - Composition of the calcined kaolinitic clay, dry basis

<u>Constituent</u>	<u>wt pct</u>
Al ₂ O ₃	41.2
SiO ₂	54.4
TiO ₂	3.1
Fe ₂ O ₃	1.3
Total.....	100.0

SULFUROUS ACID LEACHING SECTION

Calcined clay is mixed with sulfurous acid leach solution and pumped to 12 trains of 10 pressure-leach tanks. The pressure is maintained at 160 psig and the temperature at 60° C (140° F) during the 17-h leach. Sulfurous acid is injected into the tanks to maintain a 30-pct sulfurous acid solution. Calcined clay reacts with sulfurous acid to produce aluminum sulfite by the following reaction:

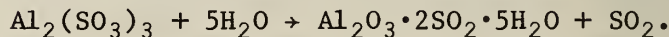


Sixty-seven percent of the alumina in the clay is assumed to react.

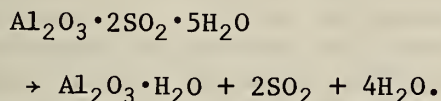
The leach slurry is pumped to pressure-leaf filters where the solids are separated from the pregnant liquor at a rate of 3.8 gal/ft²·h and washed. The filter cake is dumped from the filters into sumps, where the solids are reslurried, neutralized with slaked lime, and then pumped to a tailings pond. Filtrate and washings are combined in a surge tank and pumped to the sulfite precipitation and decomposition section. Due to the pressure drop during filtration, a portion of the sulfurous acid decomposes to form sulfur dioxide. This vapor is collected as it is blown off from the filtrate receiving vessels and sent to the sulfur dioxide handling section.

SULFITE PRECIPITATION AND DECOMPOSITION SECTION

Pregnant liquor from leaching is pumped to autoclaves that operate at 110° C (230° F) and 60 psig. Under these conditions, aluminum sulfite precipitates as monobasic aluminum sulfite by the following reaction:



Ninety-four percent of the aluminum sulfite decomposes in 2 h. Sulfur dioxide formed during the reaction is vented and sent to the sulfur dioxide handling section. The exiting hot slurry is depressurized, partially cooled by preheating the incoming solution in heat exchangers, and then pumped to a thickener. Overflow from the thickener is recycled to the sulfurous acid leach tanks. Underflow, consisting of a 19-pct monobasic aluminum sulfite slurry, is pumped to a second set of autoclaves operating at 150° C (302° F) and 50 psig. At this temperature and pressure, the monobasic aluminum sulfite will further decompose to form alumina monohydrate by the following reaction:



Decomposition requires a 1-h residence time in the reactors.

Sulfur dioxide and water vapor that are produced and vented during decomposition are collected and sent to the sulfur dioxide handling section. All sulfur dioxide process losses are included with the decomposition vapor in the material balance. Exiting hot slurry is used to preheat the autoclave feed and then pumped to rotary-vacuum drum filters. The alumina monohydrate filter cake is conveyed to the caustic digestion section. Filtrate from the drum filters is reslurried with the leach residue and pumped to the tailings pond for neutralization.

SULFUR DIOXIDE HANDLING SECTION

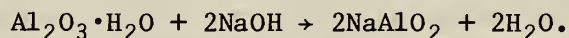
Makeup sulfur dioxide is produced by burning liquid sulfur. This process forms a hot gas containing approximately 19 pct sulfur dioxide. The gas is partially cooled in a waste-heat boiler and then passed through an absorption tower, where the sulfur dioxide is absorbed by water. This weak sulfurous acid solution is then passed through a stripping tower, where most of the water is removed to

produce a concentrated sulfur dioxide gas stream. Stripping wastes are used to reslurry the process residues for disposal in the tailings pond.

Mixed sulfur dioxide and water vapors produced in the sulfite precipitation and decomposition section are concentrated by removing some of the water through condensation. This vapor is combined with the concentrated makeup sulfur dioxide gas stream and the two sulfur dioxide-rich gas streams recovered from leaching and sulfite precipitation. The combined streams are compressed to form a strong sulfurous acid and recycled to the leach tanks.

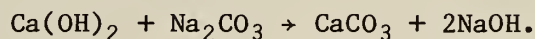
CAUSTIC DIGESTION SECTION

To remove excessive quantities of impurities from the crude decomposition product, the filter cake is processed by a modified Bayer process. Caustic solution is mixed with the alumina monohydrate filter cake and pumped into pressure-digestion vessels. Steam is injected directly into these tanks to maintain the operating condition of 230° C (446° F) and 400 psig (13). Alumina monohydrate reacts in the following manner:



Ninety-eight percent of the alumina monohydrate is dissolved during the 30-min retention time.

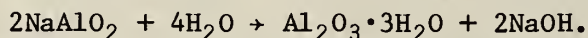
The slurry is exposed to air containing carbon dioxide in open tanks and thickeners, and by the air-lift agitators used in the precipitation tanks. This causes some of the caustic to react to form sodium carbonate. Since sodium carbonate does not dissolve alumina, lime is added in the digestion liquor regeneration section to react with the sodium carbonate and regenerate the caustic solution by forming insoluble calcium carbonate by the following reaction:



Slurry exiting the digestion tanks is depressurized and cooled in a series of nine flash tanks. Steam is recovered from the flash tanks at the pressures shown in the material balance, and used to preheat the caustic leach solution. Cooled slurry is filtered, and the solids are washed, reslurried, and pumped to the tailings pond for neutralization. Filtrate and washings are combined and pumped to the trihydrate precipitation and calcination section.

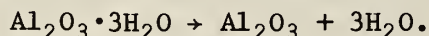
TRIHYDRATE PRECIPITATION AND CALCINATION SECTION

Pregnant solution from the caustic digestion section is pumped to 30 precipitation tanks, 30 ft in diameter by 64 ft in height, which are equipped with air lifts for agitation. Seed alumina trihydrate crystals are fed to the precipitation tanks in an amount equivalent to 100 pct of the alumina trihydrate precipitated. Seeding the supersaturated solution precipitates alumina trihydrate by the following reaction:



Forty hours are allowed for precipitation.

The slurry from precipitation is pumped to a series of three thickeners. Coarse alumina trihydrate crystals are recovered in the underflow from the primary thickener. They are filtered, washed, and fed to a fluidized-bed calciner. In the calciner, alumina trihydrate is converted to alumina by the following reaction:



This reaction occurs at 950° to 1,050° C (1,740° to 1,920° F). The cooled alumina product is conveyed to silos with a 60-day capacity for storage and shipment.

Filtrate and washings from the calciner filters are combined with overflow from the primary thickener and pumped to a secondary thickener. Overflow from this thickener is sent to a tertiary thickener. The clarified overflow from the tertiary thickener is pumped to the digestion liquor regeneration section. Underflows from both the secondary and tertiary thickeners are combined and recycled to the precipitation tanks to provide the alumina trihydrate seed crystals.

DIGESTION LIQUOR REGENERATION SECTION

Clarified solution from the tertiary thickener in the trihydrate precipitation and calcination section is concentrated in a five-effect evaporator and pumped to storage tanks. A bleed stream is pumped from the storage tank to a single-effect evaporator. Further concentration of the bleed stream allows for removal and control of minor impurities and organic compounds, which build up in the process stream. Water vapor is condensed and returned to the storage tanks along with the purified bleed stream.

Concentrated solution is pumped to a mixing tank where makeup caustic and lime are added. This solution is preheated using steam recovered from flash-cooling the digestion liquor and recycled to the digestion tanks.

Equipment has been provided for cleaning and descaling the heat exchangers used throughout the process. This equipment consists of tanks to store the solution removed from the heat exchangers, tanks containing a sulfuric acid cleaning solution, and tanks for the spent cleaning solution, as well as the necessary pumps and feed tanks.

COST ESTIMATE

The cost estimate presented in this report is based on laboratory data from the Bureau of Mines (12) and Bayer process data from published and nonpublished sources.

CAPITAL COSTS

The capital cost estimate is of the general type called a study estimate by Weaver and Bauman (14). This type of estimate, prepared from a flowsheet and a minimum of equipment data, can be expected to be within 30 pct of the actual

cost for the plant described. Although the degree of confidence in any specific study estimate is not great with respect to the actual cost, greater confidence is justified when comparing a group of similar processes evaluated by identical methods.

The estimated fixed capital cost on a fourth quarter 1982 basis (Marshall and Swift (M and S) index of 749.3) of a plant producing 1,000 tons of alumina per day is about \$752 million, as shown in table 2. This is equivalent to a cost of

TABLE 2. - Estimated capital cost¹

Fixed capital:	
Clay preparation section.....	\$46,084,000
Sulfurous acid leaching section.....	255,061,600
Sulfite precipitation and decomposition section.....	40,371,600
Sulfur dioxide handling section.....	11,008,500
Caustic digestion section.....	18,205,100
Trihydrate precipitation and calcination section.....	36,828,900
Digestion liquor regeneration section.....	24,685,600
Tailings pond.....	5,879,400
Steamplant.....	30,895,600
Subtotal.....	469,020,300
Plant facilities, 10 pct of above subtotal.....	46,902,000
Plant utilities, 12 pct of above subtotal.....	56,282,400
Total plant cost.....	572,204,700
Land cost.....	0
Subtotal.....	572,204,700
Interest during construction period.....	135,366,800
Fixed capital cost.....	<u>707,571,500</u>
Working capital:	
Raw material and supplies.....	1,283,800
Product and in-process inventory.....	13,686,000
Accounts receivable.....	13,686,000
Available cash.....	8,738,200
Working capital cost.....	37,394,000
Capitalized startup costs.....	7,075,700
Subtotal.....	<u>44,469,700</u>
Total capital cost.....	752,041,200

¹Basis: M and S equipment cost index of 749.3.

about \$2,000 per annual ton of product. The plant is designed to operate 3 shifts per day, 7 d/wk, 350 d/yr, except for some of the clay receiving facilities, which operate 2 shifts per day, 5 d/wk, and the clay crushing facilities, which operate 2 shifts per day, 7 d/wk.

Equipment costs for the proposed process are based on cost-capacity data and manufacturers' cost quotations. Cost data are brought up to date by the use of inflation indexes. Capital costs for the fluidized-bed flash calciner are based on a paper by Lussky (15). The tailings pond is designed as a lined pond with a 2-yr life. It is assumed that after 2 yr, the mine site will be developed to dispose of the remainder of the process residue as backfill. In developing the plant capital costs, corrosion-resistant materials of construction were used where appropriate. For example, the leach tanks are constructed of stainless steel to withstand the acid environment.

Factors for piping, etc., except for the foundation and electrical factors, are assigned to each section, using as a basis the effect fluids, solids, or a combination of fluids and solids may have on the process equipment. The foundation factor is estimated for each piece of equipment individually, and a factor for the entire section is calculated from the totals. The electrical factor is based on the motor horsepower requirements for each section. A factor of 10 pct, referred to as miscellaneous, is added to each section to cover minor equipment and construction costs that are not shown with the equipment listed.

For each section, the field indirect cost, which covers field supervision, inspection, temporary construction, equipment rental, and payroll overhead, is estimated at 10 pct of the direct cost. Engineering cost is estimated at 10 pct, and administration and overhead cost is estimated at 5 pct of the construction cost. A contingency allowance of 15 pct and a contractor's fee of 5 pct are included in the section costs.

The costs of plant facilities and plant utilities are estimated as 10 and 12 pct, respectively, of the total process section costs and include the same field indirect costs, engineering, administration and overhead, contingency allowance, and contractor's fee as are included in the section costs. Included under plant facilities are the costs of material and labor for auxiliary buildings such as offices, shops, laboratories, and cafeterias, and the cost of nonprocess equipment such as office furniture, and safety, shop, and laboratory equipment. Also included are labor and material costs for site preparation such as clearing, grading, drainage, roads, and fences. The costs of water, power, and steam distribution systems are included under plant utilities.

The cost for interest on the capital borrowed for construction is included as interest during construction, assuming an interest rate of 11 pct. To determine the interest cost during construction, the total plant cost is factored by an adjusted interest rate, which depends upon the length of the construction period and the interest rate at which the money is borrowed. Cost for the plant owner's supervision is not included in the capital cost of the proposed plant.

Working capital is defined as the funds in addition to fixed capital, land investment, and startup costs that must be provided to operate the plant. Working capital, also shown in table 2, is estimated from the following items: (1) Raw material and supplies inventory (cost of raw material and operating supplies for 30 days), (2) product and in-process inventory (total operating cost for 30 days), (3) accounts receivable (total operating cost for 30 days), and (4) available cash (direct expenses for 30 days).

Capitalized startup costs are estimated as 1 pct of the fixed capital, which is shown in table 2.

OPERATING COSTS

The estimated operating costs are based on the average of 350 d/yr of operation over the life of the plant. This allows 15 days' downtime per year for inspection, maintenance, and unscheduled interruptions. The operating costs are divided into direct, indirect, and fixed costs.

Direct costs include raw materials, utilities, direct labor, plant maintenance, payroll overhead, and operating supplies. The raw material costs do not include transportation costs because the plant is assumed to be located adjacent to the clay pit. Electricity, water, fuel oil, and coal are purchased utilities. The temperature of the water from the cooling tower is assumed to be 32° C (90° F). Raw material and utility requirements per ton of alumina are shown in table A-1 (appendix).

The direct labor assignments are shown by sections in table A-2. The direct labor cost is estimated on the basis of assigning 4.2 employees to each position that operates 24 h/d, 7 d/wk and 2.8 employees to each position that operates 16 h/d, 7 d/wk. The cost of labor supervision is estimated as 15 pct of the labor cost.

Plant maintenance is separately estimated for each piece of equipment and for

the buildings, electrical system, piping, plant utility distribution systems, and plant facilities.

Payroll overhead, estimated as 35 pct of direct labor and maintenance labor, includes vacation, sick leave, social security, and fringe benefits. The cost of operating supplies is estimated as 20 pct of the cost of plant maintenance.

Indirect costs are estimated as 40 pct of the direct labor and maintenance costs. The indirect costs include the expenses of control laboratories, accounting, plant protection and safety, plant administration, marketing, and company overhead. Research and overall company administrative costs outside the plant are not included.

Fixed costs include the cost of taxes (excluding income taxes), insurance, and depreciation. The annual costs of both taxes and insurance are each estimated as 1 pct of the plant construction cost. Depreciation is based on a straight-line, 20-yr period.

The estimated annual operating cost, shown in table 3, for a plant producing 1,000 tons of alumina per day is about \$166 million, or \$475 per ton of alumina product.

ECONOMIC AND TECHNICAL DISCUSSION

From the cost estimate presented in this report, the proposed process does not appear to be an economically competitive method for producing alumina. The operating cost of \$475 per ton of alumina vastly exceeds the estimated operating cost of about \$250 per ton of alumina for the Bayer process (16).

Costs for utilities, mainly electric power and coal for producing steam, account for over 30 pct of the total operating cost. Total steam requirements for this process are over 19 billion Btu/d, which is equivalent to almost

24,000 lb of steam per ton of alumina. Sulfite precipitation and decomposition are responsible for 72 pct of this steam requirement. Most of the electric power required is utilized in the leach tanks and the sulfur dioxide compressors. The slurry in the leach tanks must be continually agitated during the 17-h retention time, and with the large number of leach tanks necessary, this operation consumes an inordinate amount of electric power. Compressors require a large amount of electricity to compress over 12,000 ton/d of sulfur dioxide gas.

TABLE 3. - Estimated annual operating cost

	Annual cost	Cost per ton alumina
Direct cost:		
Raw materials:		
Kaolinitic clay at \$3 per ton.....	\$5,616,500	\$16.05
Lime at \$31.25 per ton.....	273,400	.78
Sodium hydroxide, 50-pct at \$175 per ton.....	1,041,300	2.98
Sulfur at \$114.75 per ton.....	2,128,600	6.08
Limestone at \$4 per ton.....	112,000	.32
Chemicals for steamplant water treatment.....	547,400	1.56
Total.....	9,719,200	27.77
Utilities:		
Electric power at \$0.032 per kW·h.....	12,108,100	34.59
Process water at \$0.25 per Mgal.....	668,600	1.91
Coal at \$45 per ton.....	30,835,300	88.10
Heavy oil at \$1 per gallon.....	6,634,000	18.95
Total.....	50,246,000	143.55
Direct labor:		
Labor at \$9 per hour.....	3,425,800	9.79
Supervision, 15 pct of labor.....	513,900	1.47
Total.....	3,939,700	11.26
Plant maintenance:		
Labor.....	13,408,900	38.31
Supervision, 20 pct of maintenance labor.....	2,681,800	7.66
Materials.....	13,408,900	38.31
Total.....	29,499,600	84.28
Payroll overhead, 35 pct of above payroll.....	7,010,600	20.03
Operating supplies, 20 pct of plant maintenance.....	5,899,900	16.86
Total direct cost.....	106,315,000	303.75
Indirect cost, 40 pct of direct labor and maintenance.....	13,375,700	38.22
Fixed cost:		
Taxes, 1 pct of total plant cost.....	5,722,000	16.35
Insurance, 1 pct of total plant cost.....	5,722,000	16.35
Depreciation, 20-yr life.....	35,378,600	101.08
Total operating cost.....	166,513,300	475.75

The estimated capital cost for the proposed plant is very high, resulting in high depreciation costs. In this estimate, depreciation accounts for about 20 pct of the operating costs. Basically, all the tanks and filters necessary for the process are high-capital-cost items. However, the largest equipment cost items

are the leach tanks, the pressure-leaf filters, and the sulfite precipitation tanks in decreasing order. Because of the long retention time, a large number of leach tanks is necessary. Each tank must be designed to withstand the leach pressure and be constructed of stainless steel to withstand the acid environment.

A large number of pressure-leaf filters is necessary because of the poor filtration rate of the leach residue.

Originally it had been assumed that the sulfurous acid technology could be used to produce an alternate feed for an existing Bayer plant. The use of an existing Bayer plant could reduce the capital investment required for recovering alumina from domestic clay by about 20 pct, compared with that required for a totally new plant. (This does not include the extra capital required to modify the existing Bayer plant to enable it to accept the crude alumina from the sulfurous acid plant.) Direct operating costs for a totally new plant or a modified plant will be similar. Therefore, the only operating cost advantage will be in reduced depreciation charges. Reducing depreciation costs by 20 pct is not sufficient to allow the crude alumina produced by the sulfurous acid process to compete with bauxite as feed to a Bayer plant.

The sulfurous acid leaching process was originally investigated because of the previously estimated low cost and relative ease of recovery of sulfur dioxide as a reagent, and because it would provide a feed suitable for use in a Bayer process. A suitable feed for the Bayer process has not been obtained in the

laboratory with respect to impurity levels. In previous evaluations, it was assumed that the crude alumina would be soluble at atmospheric pressure, but since it must be pressure-leached, the cost of the process has escalated.

One of the assumptions made necessary in preparing this evaluation is that an alumina product of cell-grade purity can be obtained. This has not been demonstrated in the laboratory. The small-scale studies have concentrated on the sulfurous acid leaching and sulfite precipitation and decomposition steps with only minimal investigation of the caustic digestion step in the Bayer process. Research on sulfurous acid leaching in the laboratory did not produce a product with a low enough sulfur content to meet cell-grade specifications.

In previous evaluations, the process was judged to be economically competitive with current practices. Recent Bureau of Mines research on the sulfurous acid process determined that a longer leaching time was required, and that alumina extraction was poorer than anticipated (67 pct as compared to 80 pct, which was previously assumed based on work with German clay (11)). In light of the new data, this process has become more costly and therefore does not appear economically competitive.

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APPENDIX.--UTILITY REQUIREMENTS, DIRECT LABOR REQUIREMENTS, DAILY THERMAL REQUIREMENTS, EQUIPMENT COST SUMMARIES, AND MATERIAL BALANCES

Raw material and utility requirements per ton of alumina are shown in table A-1, and direct labor requirements and daily thermal requirements for each section are shown in tables A-2 and A-3, respectively. Major items of equipment for each section are shown in table A-4. The equipment cost summaries for each section in the process are contained in

tables A-5 to A-11. Material balances are shown for each section in figures A-1 to A-7.

TABLE A-2. - Direct labor requirements, operators per shift

TABLE A-1. - Raw material and utility requirements

	<u>Quantity</u> <u>per ton</u> <u>alumina</u>
Raw materials, tons:	
Kaolinitic clay.....	5.349
Lime.....	.025
Sodium hydroxide, 50-pct...	.017
Sulfur.....	.053
Limestone.....	.080
Utilities:	
Electric power.....kW·h..	1,081.085
Process water.....Mgal..	7.641
Coal.....tons..	1.958
Heavy oil.....gal..	18.954

Section	Shifts per week		
	¹ 21	² 14	³ 5
Clay preparation section	4	2	0
Sulfurous acid leaching section.....	14	0	0
Sulfite precipitation and decomposition section.....	4	0	0
Sulfur dioxide handling section.....	1	0	0
Caustic digestion section.....	3	0	0
Trihydrate precipitation and calcination section	2	0	1
Digestion liquor regeneration section.....	2	0	0
Steamplant.....	11	0	3
General plant.....	0	0	1
Total.....	41	2	5

¹3 shifts per day, 7 d/wk.

²2 shifts per day, 7 d/wk.

³1 shift per day, 5 d/wk.

TABLE A-3. - Daily thermal requirements

Section and item	Steam, MMBtu	Heavy oil, MMBtu ¹	Coal, MMBtu ²	Cooling water, Mgal
Clay preparation section:				
Fluidized-bed calciners.....	0	0	9,870	0
Lime scrubber.....	128	0	0	0
Total.....	128	0	9,870	0
Sulfurous acid leaching section: Leach tanks.....	0	0	0	65,971
Sulfite precipitation and decomposition section:				
Precipitation tanks.....	8,295	0	0	0
Decomposition tanks.....	5,644	0	0	0
Total.....	13,939	0	0	0
Sulfur dioxide handling section:				
Waste-heat boiler.....	-295	0	0	0
Reboiler.....	41	0	0	0
Condenser.....	0	0	0	14,135
Total.....	-254	0	0	14,135
Caustic digestion section:				
Heat exchanger.....	751	0	0	0
Heat exchanger.....	845	0	0	0
Digestion tanks.....	2,207	0	0	0
Flash tank 1.....	-495	0	0	0
Flash tank 2.....	-437	0	0	0
Flash tank 3.....	-510	0	0	0
Flash tank 4.....	-961	0	0	0
Flash tank 5.....	-999	0	0	0
Flash tank 6.....	-560	0	0	0
Flash tank 7.....	-525	0	0	0
Flash tank 8.....	-513	0	0	0
Flash tank 9.....	-349	0	0	0
Total.....	-1,546	0	0	0
Trihydrate precipitation and calcination section: Fluid-flash calciner.....	0	2,900	0	0
Digestion liquor regeneration section:				
Multieffect evaporator.....	1,549	0	0	3,379
Evaporator.....	182	0	0	408
Heat exchanger.....	437	0	0	0
Heat exchanger.....	495	0	0	0
Heat exchanger.....	349	0	0	0
Heat exchanger.....	513	0	0	0
Heat exchanger.....	525	0	0	0
Heat exchanger.....	560	0	0	0
Heat exchanger.....	999	0	0	0
Heat exchanger.....	961	0	0	0
Heat exchanger.....	510	0	0	0
Lime slaker.....	0	0	0	89
Total.....	7,080	0	0	3,876
General plant.....	0	0	40	0
Total process requirements.....	19,347	2,900	9,910	83,982
Steamplant.....	-19,347	0	39,035	0
Total.....	0	2,900	48,945	3,982

¹Quantity of heavy oil (153,000 Btu/gal) = 18,954 gal/d.²Quantity of coal (12,500 Btu/lb) = 1,958 ton/d.

TABLE A-4. - Major items of equipment

Section and item	Quantity	Unit size
Clay preparation section:		
Hammer mills.....	3	60 by 48 in.
Pelletizing disks.....	8	12-ft diam.
Fluidized-bed calciners.....	4	1,500 ton/d.
Sulfurous acid leaching section:		
Leach tanks.....	132	15-ft diam by 45 ft.
Pressure-leaf filters.....	120	1,645 ft ² .
Sulfite precipitation and decomposition section:		
Precipitation tanks.....	16	15-ft diam by 45 ft.
Thickener.....	1	76-ft diam.
Decomposition tanks.....	5	15-ft diam by 45 ft.
Sulfur dioxide handling section:		
Absorption tower.....	1	5.9-ft diam by 22.7 ft.
Stripping tower.....	1	4.6-ft diam by 19.8 ft.
Caustic digestion section:		
Digestion tanks.....	4	12-ft diam by 36 ft.
Pressure-leaf filters.....	12	1,644 ft ² .
Trihydrate precipitation and calcination section:		
Precipitation tanks.....	36	30-ft diam by 64 ft.
Primary thickeners.....	4	27-ft diam.
Secondary thickeners.....	4	59-ft diam.
Tertiary thickeners.....	4	115-ft diam.
Fluid-flash calciner.....	1	2.9 billion Btu/d.
Digestion liquor regeneration section:		
Multieffect evaporator.....	1	15,699 ft ² /effect.
Evaporator.....	1	1,519 ft ² /effect.

TABLE A-5.-Equipment cost summary, clay preparation section

Item	Cost(1)		
	Equipment	Labor	Total
Apron feeder.....	\$ 20800.	\$ 3100.	\$ 23900.
Belt conveyor.....	67100.	13300.	80400.
Belt conveyor.....	174600.	33400.	208000.
Reclaimer feeders.....	31300.	4700.	36000.
Belt conveyor.....	198400.	35800.	234200.
Hoppers.....	9700.	800.	10500.
Apron feeders.....	23500.	3500.	27000.
Hammer mills.....	512600.	71800.	584400.
Belt conveyor.....	13300.	2500.	15800.
Belt conveyor.....	63400.	12700.	76100.
Storage bin.....	226100.	44500.	270600.
Apron feeder.....	17400.	2600.	20000.
Belt conveyor.....	46700.	6600.	53300.
Pelletizing disks.....	601200.	3600.	604800.
Screw feeders.....	12100.	1800.	13900.
Coal preparation equipment.....	382300.	34900.	417200.
Bag dust collectors.....	31600.	500.	32100.
Screw feeders.....	8700.	1300.	10000.
Total.....	2440800.	277400.	2718200.
Unloading hopper.....			(2) 17900.
Fluidized-bed calciners.....			(2) 20931300.
Lime scrubber.....			(2) 2938600.
Front-end loaders.....			136200.
Total equipment cost x factor indicated:			
Foundations, x .157.....			382000.
Buildings, x .563.....			1375000.
Structures, x .100.....			244100.
Instrumentation, x .080.....			195300.
Electrical, x .184.....			450100.
Piping, x .200.....			488200.
Painting, x .020.....			48800.
Miscellaneous, x .100.....			244100.
Total.....			3427600.
Total direct cost.....			30169800.
Field indirect, 10.0 pct of total direct cost.....			3017000.
Total construction cost.....			33186800.
Engineering, 10.0 pct of total construction cost.....			3318700.
Administration and overhead, 5.0 pct of total construction cost.....			1659300.
Subtotal.....			38164800.
Contingency, 15.0 pct of above subtotal.....			5724700.
Subtotal.....			43889500.
Contractor's fee, 5.0 pct of above subtotal.....			2194500.
Section cost.....			46084000.

(1) Equipment costs are based on the M and S index of 749.3.

(2) Installed cost.

TABLE A-6.-Equipment cost summary, sulfurous acid leaching section

Item	Cost(1)		Total
	Equipment	Labor	
Belt conveyor.....	\$ 12300.	\$ 2200.	\$ 14500.
Belt conveyor.....	122000.	22300.	144300.
Storage bins.....	575600.	117000.	692600.
Screw feeders.....	37200.	5600.	42800.
Mixing tanks.....	172000.	34900.	206900.
Pumps.....	261500.	34100.	295600.
Leach tanks.....	73939700.	184300.	74124000.
Pressure-leaf filters.....	5207500.	312900.	5520400.
Pumps.....	67800.	23600.	91400.
Sumps.....	210700.	44900.	255600.
Pumps.....	155000.	35900.	190900.
Hoppers.....	19800.	6800.	26600.
Belt conveyors.....	281100.	65400.	346500.
Total.....	81062200.	889900.	81952100.
Total equipment cost x factor indicated:			
Foundations, x .049.....			3944100.
Buildings, x .010.....			774600.
Structures, x .100.....			8106200.
Insulation, x .011.....			895000.
Instrumentation, x .180.....			14591200.
Electrical, x .020.....			1595600.
Piping, x .500.....			40531100.
Painting, x .080.....			6485000.
Miscellaneous, x .100.....			8106200.
Total.....			85029000.
Total direct cost.....			166981100.
Field indirect, 10.0 pct of total direct cost.....			16698100.
Total construction cost.....			183679200.
Engineering, 10.0 pct of total construction cost.....			18367900.
Administration and overhead, 5.0 pct of total construction cost.....			9184000.
Subtotal.....			211231100.
Contingency, 15.0 pct of above subtotal.....			31684700.
Subtotal.....			242915800.
Contractor's fee, 5.0 pct of above subtotal.....			12145800.
Section cost.....			255061600.

(1) Equipment costs are based on the M and S index of 749.3.

TABLE A-7.-Equipment cost summary,
sulfite precipitation and decomposition section

Item	Cost(1)		
	Equipment	Labor	Total
Precipitation tanks.....	\$ 3153600.	\$ 22300.	\$ 3175900.
Blowdown tanks.....	46300.	23700.	70000.
Heat exchangers.....	1003200.	31400.	1034600.
Thickener.....	459500.	30000.	489500.
Overflow pumps.....	24100.	4700.	28800.
Underflow pumps.....	39100.	5000.	44100.
Decomposition tanks.....	912300.	7000.	919300.
Blowdown tanks.....	21200.	9500.	30700.
Heat exchangers.....	262900.	8700.	271600.
Pumps.....	62400.	11900.	74300.
Rotary-vacuum drum filters.....	4443200.	222200.	4665400.
Pumps.....	98700.	26300.	125000.
Reslurry tanks.....	121300.	23500.	144800.
Pumps.....	102900.	21000.	123900.
Sulfur dioxide storage tank.....	319400.	1100.	320500.
Total.....	11070100.	448300.	11518400.
Cooling tower.....			(2) 1886400.
Total equipment cost x factor indicated:			
Foundations, x .093.....			1024300.
Buildings, x .029.....			318200.
Structures, x .100.....			1107000.
Insulation, x .062.....			684400.
Instrumentation, x .180.....			1992600.
Electrical, x .034.....			371000.
Piping, x .500.....			5535100.
Painting, x .080.....			885600.
Miscellaneous, x .100.....			1107000.
Total.....			13025200.
Total direct cost.....			26430000.
Field indirect, 10.0 pct of total direct cost.....			2643000.
Total construction cost.....			29073000.
Engineering, 10.0 pct of total construction cost.....			2907300.
Administration and overhead, 5.0 pct of total construction cost.....			1453700.
Subtotal.....			33434000.
Contingency, 15.0 pct of above subtotal.....			5015100.
Subtotal.....			38449100.
Contractor's fee, 5.0 pct of above subtotal.....			1922500.
Section cost.....			40371600.

(1) Equipment costs are based on the M and S index of 749.3.

(2) Installed cost.

TABLE A-8.-Equipment cost summary, sulfur dioxide handling section

Item	Cost(1)		
	Equipment	Labor	Total
Liquid sulfur storage tank.....	\$ 65900.	\$ 32000.	\$ 97900.
Pumps.....	14800.	1400.	16200.
Sulfur furnace.....	9100.	14300.	23400.
Waste-heat boiler.....	47400.	1500.	48900.
Blowers.....	117100.	700.	117800.
Absorption tower.....	41900.	800.	42700.
Pumps.....	12800.	3200.	16000.
Pumps.....	12800.	3200.	16000.
Stripping tower.....	27900.	500.	28400.
Reboiler.....	4600.	300.	4900.
Pumps.....	7900.	2200.	10100.
Condenser.....	66300.	1700.	68000.
Pumps.....	16300.	3900.	20200.
Compressors.....	2578000.	22800.	2600800.
Total.....	3022800.	88500.	3111300.
Water chiller.....			(2) 50900.
Total equipment cost x factor indicated:			
Foundations, x .078.....			234800.
Buildings, x .011.....			34100.
Structures, x .200.....			604600.
Insulation, x .031.....			93300.
Instrumentation, x .080.....			241800.
Electrical, x .228.....			689900.
Piping, x .600.....			1813700.
Painting, x .010.....			30200.
Miscellaneous, x .100.....			302300.
Total.....			4044700.
Total direct cost.....			7206900.
Field indirect, 10.0 pct of total direct cost.....			720700.
Total construction cost.....			7927600.
Engineering, 10.0 pct of total construction cost.....			792800.
Administration and overhead, 5.0 pct of total construction cost.....			396400.
Subtotal.....			9116800.
Contingency, 15.0 pct of above subtotal.....			1367500.
Subtotal.....			10484300.
Contractor's fee, 5.0 pct of above subtotal.....			524200.
Section cost.....			11008500.

(1) Equipment costs are based on the M and S index of 749.3.

(2) Installed cost.

TABLE A-9.-Equipment cost summary, caustic digestion section

Item	Cost(1)		
	Equipment	Labor	Total
Belt conveyors.....	\$ 568700.	\$ 142600.	\$ 711300.
Hopper.....	5400.	900.	6300.
Screw feeder.....	13600.	700.	14300.
Mixing tank.....	29400.	5500.	34900.
Pumps.....	22900.	4300.	27200.
Pumps.....	21400.	4300.	25700.
Heat exchanger.....	21000.	1100.	22100.
Heat exchanger.....	27500.	1300.	28800.
Pumps.....	146100.	9200.	155300.
Digestion tanks.....	2747200.	4600.	2751800.
Flash tank 1.....	73000.	800.	73800.
Flash tank 2.....	55800.	800.	56600.
Flash tank 3.....	44200.	800.	45000.
Flash tank 4.....	35000.	800.	35800.
Flash tank 5.....	33400.	800.	34200.
Flash tank 6.....	33400.	800.	34200.
Flash tank 7.....	33400.	800.	34200.
Flash tank 8.....	33400.	800.	34200.
Flash tank 9.....	33400.	800.	34200.
Pumps.....	27900.	4200.	32100.
Pressure-leaf filters.....	356700.	35700.	392400.
Reslurry tanks.....	80800.	24800.	105600.
Pumps.....	101400.	18800.	120200.
Total.....	4545000.	265200.	4810200.
Slurry storage tank.....			(2) 355200.
Total equipment cost x factor indicated:			
Foundations, x .211.....			958600.
Buildings, x .041.....			186600.
Structures, x .100.....			454500.
Insulation, x .017.....			79300.
Instrumentation, x .180.....			818100.
Electrical, x .056.....			256300.
Piping, x .700.....			3181500.
Painting, x .080.....			363600.
Miscellaneous, x .100.....			454500.
Total.....			6753000.
Total direct cost.....			11918400.
Field indirect, 10.0 pct of total direct cost.....			1191800.
Total construction cost.....			13110200.
Engineering, 10.0 pct of total construction cost.....			1311000.
Administration and overhead, 5.0 pct of total construction cost.....			655500.
Subtotal.....			15076700.
Contingency, 15.0 pct of above subtotal.....			2261500.
Subtotal.....			17338200.
Contractor's fee, 5.0 pct of above subtotal.....			866900.
Section cost.....			18205100.

(1) Equipment costs are based on the M and S index of 749.3.

(2) Installed cost.

TABLE A-10.-Equipment cost summary,
trihydrate precipitation and calcination section

Item	Cost(1)		
	Equipment	Labor	Total
Pumps.....	\$ 62200.	\$ 24400.	\$ 86600.
Precipitation tanks.....	1740100.	680800.	2420900.
Pumps.....	226000.	62800.	288800.
Primary thickeners.....	338300.	49900.	388200.
Sumps.....	4100.	5300.	9400.
Underflow pumps.....	29500.	5700.	35200.
Secondary thickeners.....	644400.	96800.	741200.
Sumps.....	20600.	14500.	35100.
Underflow pumps.....	52900.	9600.	62500.
Tertiary thickeners.....	1578200.	170500.	1748700.
Sumps.....	5700.	6500.	12200.
Underflow pumps.....	24700.	5400.	30100.
Surge tanks.....	29600.	11800.	41400.
Pumps.....	61600.	10000.	71600.
Overflow pumps.....	41800.	9200.	51000.
Surge tanks.....	25400.	16400.	41800.
Pumps.....	41800.	9200.	51000.
Belt conveyor.....	146700.	28700.	175400.
Screw feeders.....	32500.	4900.	37400.
Total.....	5106100.	1222400.	6328500.
Air compressors.....			(2) 223400.
Fluid-flash calciner.....			(2) 6148600.
Silos.....			(2) 4817300.
Total equipment cost x factor indicated:			
Foundations, x .429.....			2191200.
Buildings, x .025.....			130100.
Structures, x .050.....			255300.
Instrumentation, x .150.....			765900.
Electrical, x .087.....			442100.
Piping, x .400.....			2042400.
Painting, x .050.....			255300.
Miscellaneous, x .100.....			510600.
Total.....			6592900.
Total direct cost.....			24110700.
Field indirect, 10.0 pct of total direct cost.....			2411100.
Total construction cost.....			26521800.
Engineering, 10.0 pct of total construction cost.....			2652200.
Administration and overhead, 5.0 pct of total construction cost.....			1326100.
Subtotal.....			30500100.
Contingency, 15.0 pct of above subtotal.....			4575000.
Subtotal.....			35075100.
Contractor's fee, 5.0 pct of above subtotal.....			1753800.
Section cost.....			36828900.

(1) Equipment costs are based on the M and S index of 749.3.

(2) Installed cost.

TABLE A-11.-Equipment cost summary,
digestion liquor regeneration section

Item	Cost(1)		
	Equipment	Labor	Total
Multieffect evaporator.....	\$ 3067000.	\$ 300800.	\$ 3367800.
Pumps.....	12200.	3100.	15300.
Condensate tank.....	23300.	8000.	31300.
Pumps.....	10000.	2700.	12700.
Pumps.....	17600.	4100.	21700.
Pumps.....	4900.	1200.	6100.
Surge tank.....	1500.	1600.	3100.
Pumps.....	4900.	1200.	6100.
Evaporator.....	234500.	47500.	282000.
Condensate tank.....	200.	400.	600.
Pumps.....	4900.	800.	5700.
Pumps.....	4900.	900.	5800.
Concentrated caustic tanks.....	65000.	23300.	88300.
Pumps.....	31000.	8700.	39700.
Pumps.....	32700.	8700.	41400.
Heat exchanger.....	39400.	1100.	40500.
Heat exchanger.....	53800.	1400.	55200.
Heat exchanger.....	56000.	1400.	57400.
Heat exchanger.....	60700.	1500.	62200.
Heat exchanger.....	109000.	2300.	111300.
Heat exchanger.....	121900.	2600.	124500.
Heat exchanger.....	97900.	1900.	99800.
Heat exchanger.....	101500.	2000.	103500.
Heat exchanger.....	139300.	2400.	141700.
Pumps.....	54100.	17000.	71100.
Solution storage tanks.....	415200.	152900.	568100.
Acid storage tanks.....	12100.	8800.	20900.
Pumps.....	14800.	3900.	18700.
Feed tanks.....	1700.	2600.	4300.
Pumps.....	15300.	3900.	19200.
Spent acid storage tanks.....	12100.	8800.	20900.
Pumps.....	15300.	3900.	19200.
Pumps.....	9900.	500.	10400.
Makeup storage tank.....	23700.	8800.	32500.
Pumps.....	4900.	400.	5300.
Caustic dilution tanks.....	14800.	1300.	16100.
Pumps.....	4900.	500.	5400.
Lime receivers.....	247800.	129200.	377000.
Screw feeders.....	10900.	1600.	12500.
Belt conveyor.....	36300.	5700.	42000.
Storage bin.....	4100.	1900.	6000.
Screw feeder.....	2200.	300.	2500.
Pumps.....	4900.	900.	5800.
Total.....	5199100.	782500.	5981600.
Caustic storage tanks.....			(2) 1643800.
Lime slaker.....			(2) 55700.

TABLE A-11.-Equipment cost summary,
digestion liquor regeneration section
(continued)

Total equipment cost x factor indicated:	
Foundations, x .345.....	1794200.
Buildings, x .015.....	77500.
Structures, x .050.....	260000.
Insulation, x .289.....	1501300.
Instrumentation, x .150.....	779900.
Electrical, x .032.....	167500.
Piping, x .600.....	3119500.
Painting, x .050.....	260000.
Miscellaneous, x .100.....	519900.
Total.....	<u>8479800.</u>
Total direct cost.....	16160900.
Field indirect, 10.0 pct of total direct cost.....	1616100.
Total construction cost.....	<u>17777000.</u>
Engineering, 10.0 pct of total construction cost.....	1777700.
Administration and overhead, 5.0 pct of total construction cost.....	888900.
Subtotal.....	<u>20443600.</u>
Contingency, 15.0 pct of above subtotal.....	3066500.
Subtotal.....	<u>23510100.</u>
Contractor's fee, 5.0 pct of above subtotal.....	1175500.
Section cost.....	<u>24685600.</u>

- (1) Equipment costs are based on the M and S index of 749.3.
(2) Installed cost.

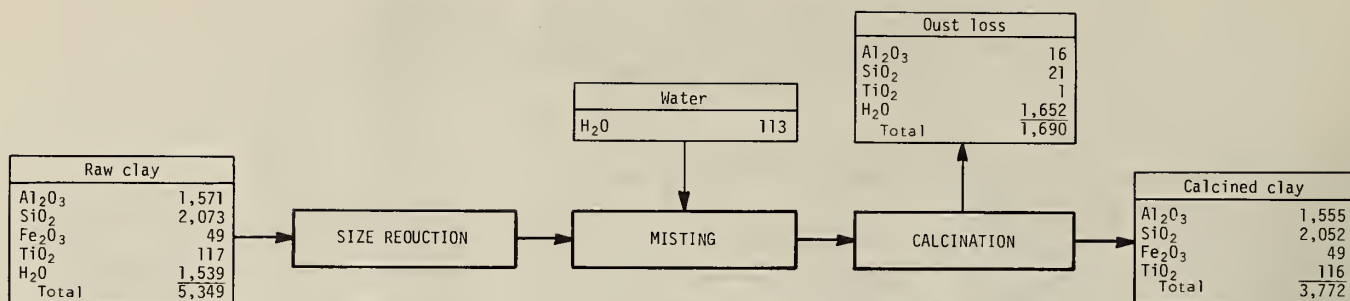


FIGURE A-1. - Material balance, clay preparation section (tons per day).

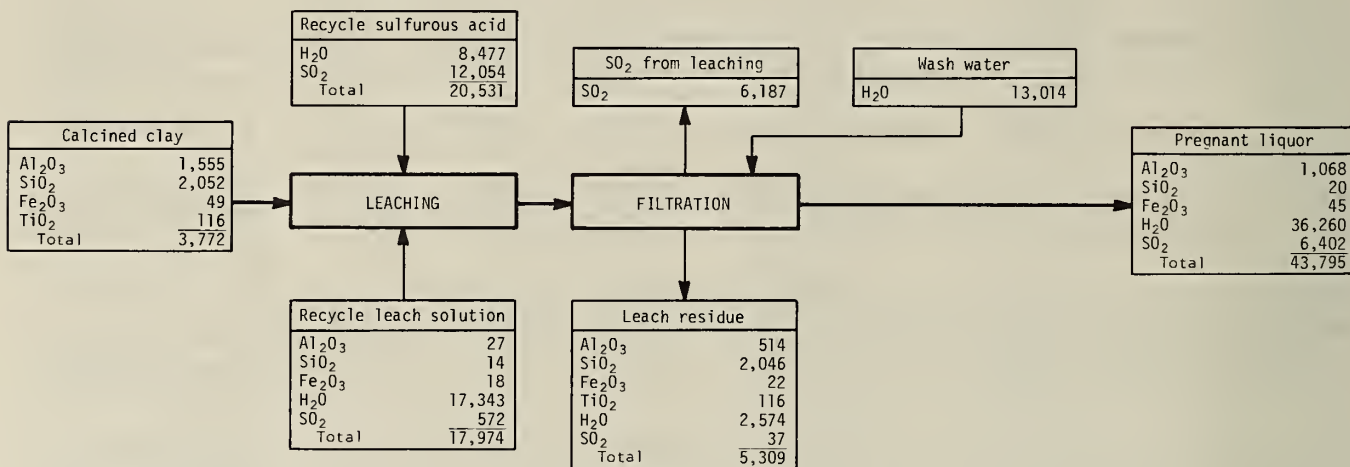


FIGURE A-2. - Material balance, sulfurous acid leaching section (tons per day).

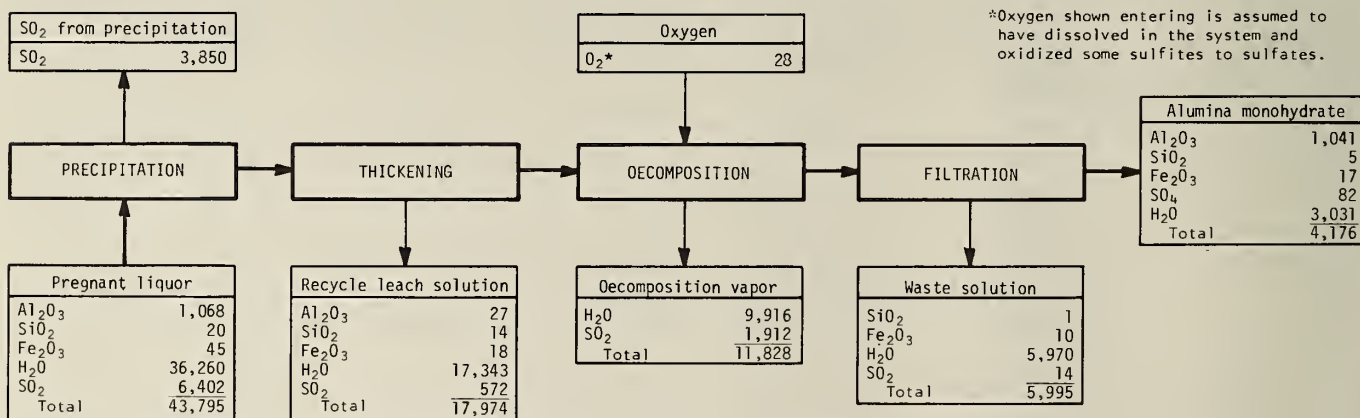


FIGURE A-3. - Material balance, sulfite precipitation and decomposition section (tons per day).

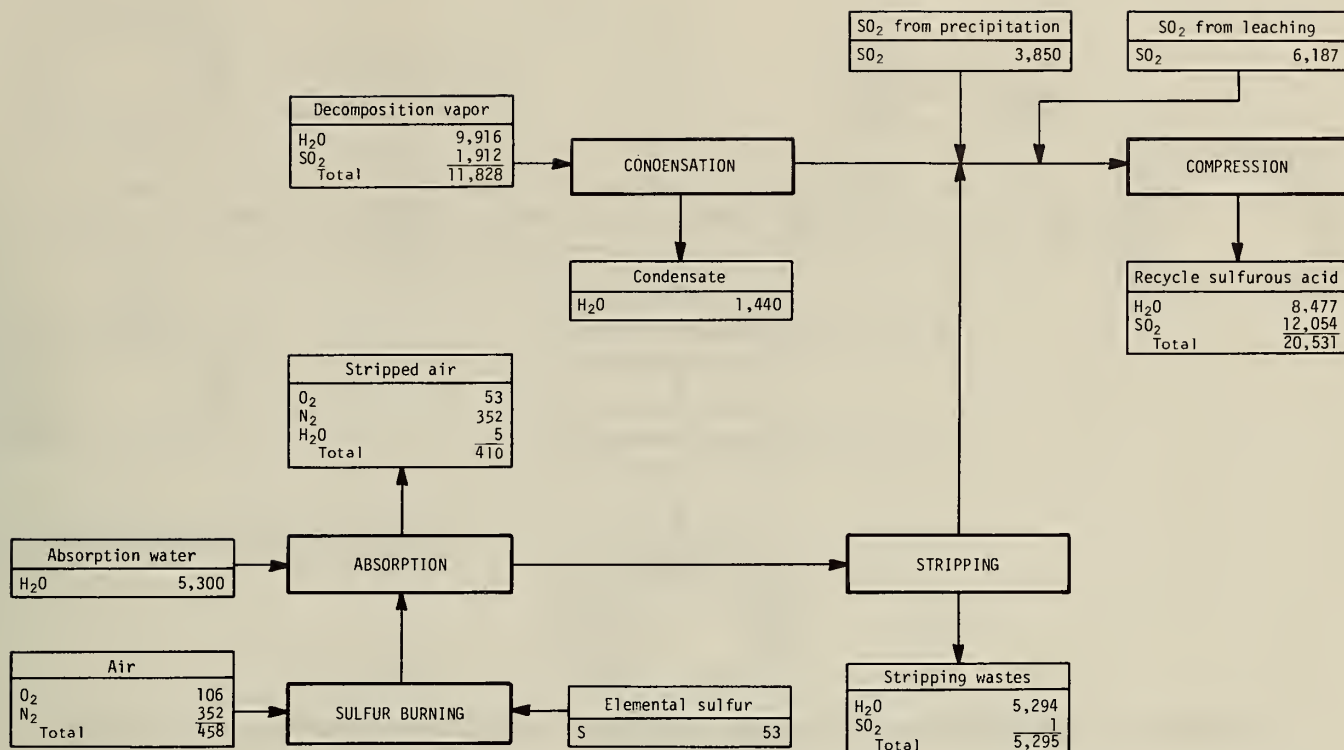


FIGURE A-4. - Material balance, sulfur dioxide handling section (tons per day).

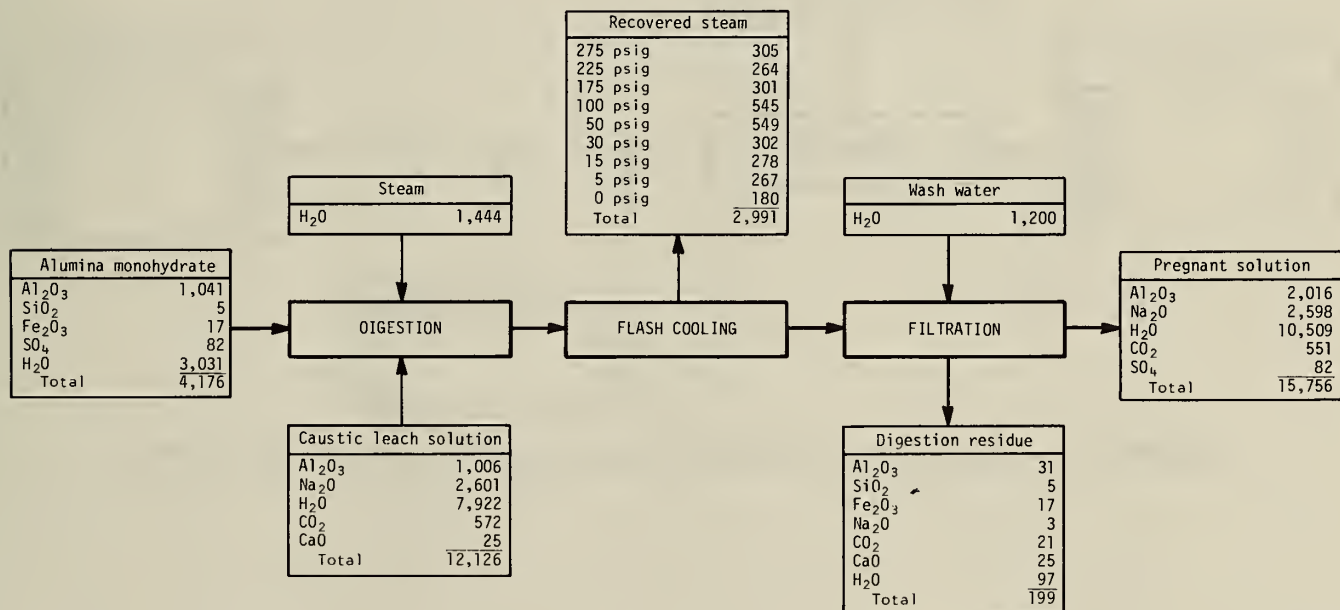


FIGURE A-5. - Material balance, caustic digestion section (tons per day).

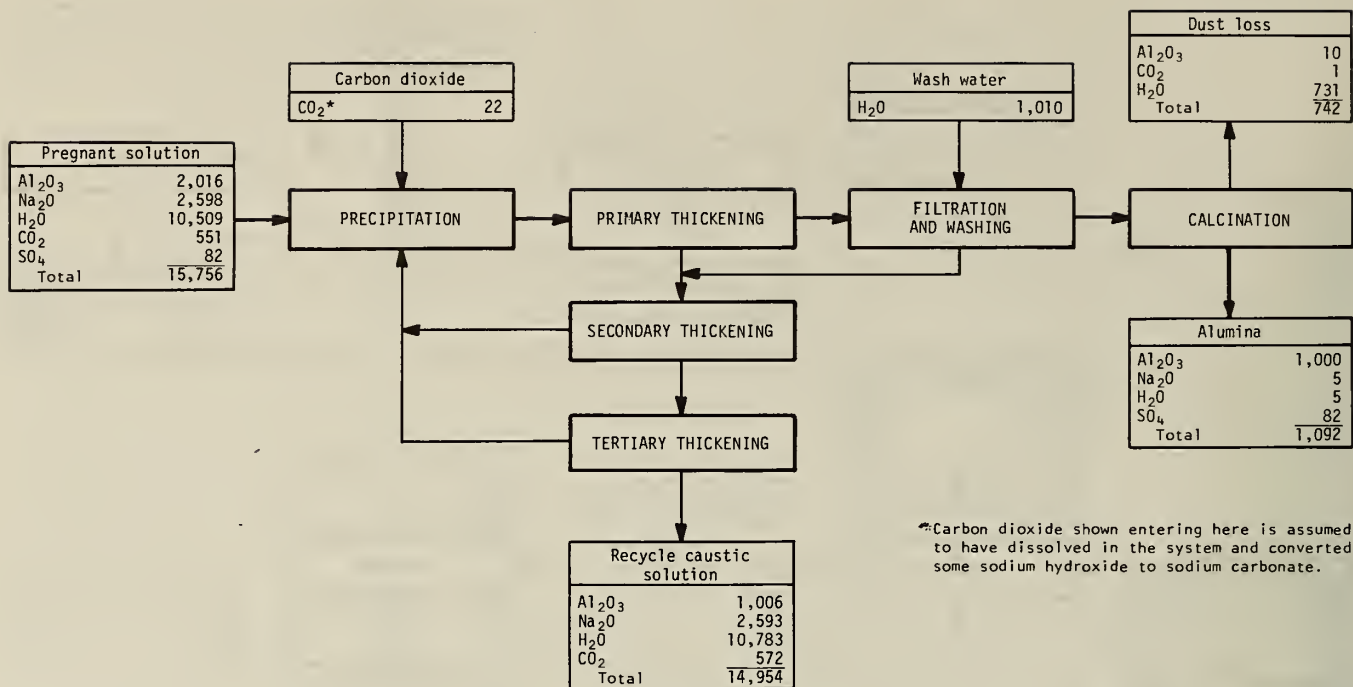


FIGURE A-6. - Material balance, trihydrate precipitation and calcination section (tons per day).

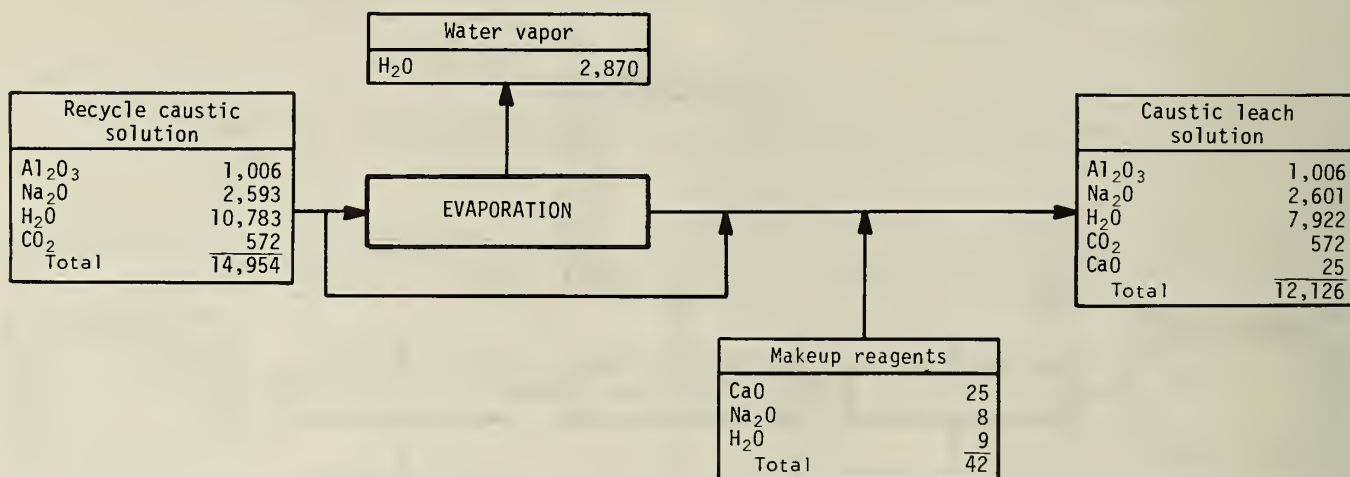


FIGURE A-7. - Material balance, digestion liquor regeneration section (tons per day).

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